

## BIODIESEL PRODUCTION PROCESS FLOW DESIGN VIA CALCIUM OXIDE CATALYSED ETHANOLYSIS TECHNIQUE

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### ABSTRACT

*This paper presents a report on the design of a process flow method and a case study of a mini-waste cooking oil biodiesel (WCOBD) plant. This report comprises a detailed process development, technical and economic assessment incorporating material and energy computations, equipment cost estimations, and specifications. The designed plant is expected to produce 16.88 tons per annum. The cost of biodiesel was evaluated based on researched cost variables and plant data. The designed plant seems economically viable, and a biodiesel production cost of ZAR10.10 per kg was evaluated by the investigation giving a total production cost of ZAR10,100.00 (US\$673.33) per ton. Mass balances were utilized to design a 50 L pressurized reactor tank which is suitable for small to medium sized biodiesel producers.*

**KEYWORDS:** Waste Cooking Oil Biodiesel; Waste Cooking Oil; Cetane Number; Free Fatty Acid & FATTY Acid Alkyl Esters

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### NOMENCLATURE

ASTM	American Society for	CV	Calorific value
Testing and Materials		FAAE	Fatty acid alkyl esters
CaO Calcium Oxide		FD	Fossil diesel
(Ca (NO <sub>3</sub> ) <sub>2</sub> ·4H <sub>2</sub> O)	calcium nitrate tetrahydrate	NaOH	Sodium hydroxide
EtOH	Ethanol	NO <sub>x</sub>	Nitrogen oxide
FFA	Free fatty acid	US \$	United States Dollars
H <sub>3</sub> PO <sub>4</sub>	Phosphoric acid	WCO	Waste Cooking Oil
CN	Cetane number	WCOBD	Waste Cooking Oil
CI	Compression ignition	Biodiesel	
CO	Carbon monoxide	ZAR	South African Rand
CO <sub>2</sub>	Carbon dioxide	EAC	Equipment acquisition cost

### 1. INTRODUCTION

Massive reliance on fuel has impacted negatively on climate and the environment (Perera 2018). This scenario has

been caused by the emission of toxic gases generated when fossil fuels are burnt inside internal combustion engines which are the transport sector's major sources of energy. These have been the main drivers of global warming and climate change. This scenario has encouraged a search for renewable fuels which are non-toxic, and biodiesel has been identified as a viable replacement for fossil diesel fuel. Biodiesel is a long chain methyl ester derived from transesterification of vegetable oils or animal fats. These energy resources possess similar fuel properties to those of fossil diesel fuel. Despite gaining global attention, biodiesel is yet to gain significant market share due to the high cost of feedstock acquisition, which is why waste cooking oil (WCO) is a promising option.

Homogeneous catalyzed reactions are frequently used to transform free fatty acids (FFA) via methanol to biodiesel. However, this process has proven to be rigorous, costly, and harmful to the environment (Abdullah, Hanapi et al. 2017). The recuperation of the catalyst from the reaction byproducts is complex and challenging. Moreover, a tremendous measure of glycerol is created, and filtering the products yields a lot of wastewater, raising the cost of the purification process. Furthermore, biodiesel produced with this technique has proven to be insignificant when evaluated economically against fossil diesel. This is chiefly on the grounds that the catalyst is sensitive to high FFA containing feedstocks and can only handle feedstocks with FFA under 0.5% wt %. These high-quality feedstocks are acquired at high cost rendering an expensive fuel product. Several studies have researched the potential of other options which can possibly be more financially viable. These must incorporate utilization of less expensive feedstock and easily attainable catalysts as technological options. This specific technique must utilize less expensive catalysts with somewhat higher reactant capability, and requires simple catalytic recuperation, which enables repeated utilization of the catalysts. Heterogeneous catalysts have the capability of avoiding the drawbacks shown by homogenous base catalysis techniques. Heterogeneous catalysts have stood out as most appropriate among promising catalysts because of their high capacity to catalyze transesterification reactions and because of their reusability behavior. Besides being insensitive to feedstock FFA content, these catalysts are derived from locally accessible resources such as clay, eggshells, and so forth.

Amongst the heterogenous catalysts studied, calcium oxide (CaO) has risen to be the most fascinating option available. CaO has risen to be the most broadly researched heterogeneous catalysts among the metal oxides, capable of delivering around 96 % biodiesel yield (Ljupković, Mičić et al. 2014). This catalyst is nano crystal in nature which gives it a high surface area to volume ratio. In addition, CaO is easy to manufacture and can be developed from locally available resources such as eggshells, limestone and seashells, making it cheaper to produce compared to NaOH and KOH. Ljupković, Mičić et al. (2014) investigated the properties of CaO catalysts in the creation of biodiesel and found that these catalysts possess unique features capable of significantly improving the process. These unique properties include a shorter reaction duration (up to 120 minutes), standard working temperature (64 °C) and pressures, a generally low molar proportion (6:1), and catalysts concentration (1 wt.%). This investigation found that all the above variables enable the development of high-quality biodiesel reaching purity as high as 96 % (Ljupković, Mičić et al. 2014). A study by Avhad, Sánchez et al. (2016) revealed that ideal conditions for CaO catalyzed ethanolysis were 120 minutes reaction time, reaction temperature of 75 °C, molar ratio of 9:1 and catalyst concentration of 7 wt %.

The significant challenges related to the manufacture of biodiesel as a fuel are its high cost of creation, having generally less energy content when evaluated against fossil diesel, and delivering higher nitrogen oxide discharges when it is burnt (Avhad, Sánchez et al. 2016), with the main challenge being cost (Marchett et al., 2008). There are three ways to achieve unit cost decreases: improving the yield, diminishing capital expenses, and decreasing the crude material costs

with feedstock cost being the most significant cost (Mulugetta 2009). All these potential ways require investigation with different technologies encompassing various catalysts and feedstocks. In addition, localized engineering design of biodiesel plants is non-existent as most researchers are focusing on the chemistry side of the technologies. No studies on process design and evaluation using a batch reactor and ethanolysis catalyzed by CaO have been performed to date. Therefore, this research is aimed at designing a process flow for WCO biodiesel production from a CaO catalyzed ethanolysis process with an economic assessment of the selected system. The information acquired from this research will be used to set up a biodiesel plant in South Africa and the technology will be applicable to most of the sub-Saharan region.

## 2. PROCESS DESIGN AND CASE STUDY

### 2.1. METHODS AND MATERIALS

#### 2.1.1. Materials

Waste cooking oil was purchased from Enfield at ZAR3.25/kg and its properties were evaluated as shown in Table 1. glycerol (GL), phosphoric acid CAS number 7662-29-3, molar ratio 97.994 g/mol and 85 % purity were purchased from Corebit (Pty) Ltd at ZAR3/kg and ZAR 5.45/ kg respectively. Magnesium silicate ( $\text{MgSiO}_3$ ) CAS (1343-88-0) at ZAR 728/kg was purchased from Sigma Aldrich. Chemicals for nano CaO production were purchased from Sigma Aldrich, namely, calcium nitrate tetrahydrate ( $\text{Ca}(\text{NO}_3)_2 \cdot 4\text{H}_2\text{O}$ ) 13477-34-4 and ethylene glycol ( $\text{C}_2\text{H}_6\text{O}_2$ ) CAS number: 107-21-1, sodium hydroxide (NaOH) CAS number: 1310-73-2 at ZAR 400, ZAR12.00 and ZAR6.00 respectively. The safety precautions described for each chemical reagents from the suppliers were followed during the experimental process.

**Table 1: Waste Cooking Oil Properties**

Oil Property	Values	ASTM Limits*	Units
Flash point	168	>130	°C
Density (40 °C)	890		$\text{Kg/m}^3$
Kinematic viscosity	3.9	1.9–6.0	$\text{mm}^2/\text{s}$
Water and sediments	0.04	<0.05	%
Acid number	2.56	0.50 max	mg KOH/g
Free glycerol	0.013	0.020	wt. %
Total glycerol	0.018	0.024	wt. %

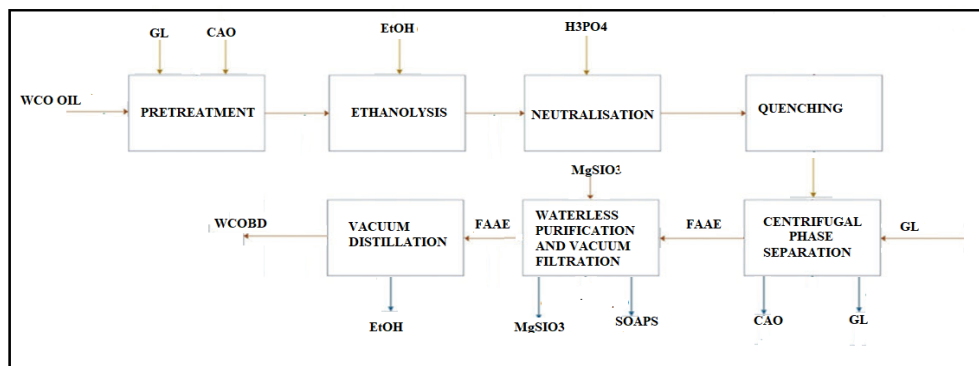
#### 2.1.2. CaO nano- Catalysts Synthesis

0.74 kg of CaO catalyst was developed using the thermal decomposition technique as per Tang, Claveau et al. (2008). 3.20 kg of calcium nitrate tetrahydrate ( $\text{Ca}(\text{NO}_3)_2 \cdot 4\text{H}_2\text{O}$ ) was dissolved in a 6.43 ml of ethylene glycol solution. Thereafter 0.57 kg of NaOH was added. The mixture was thoroughly stirred for 12 min and thereafter kept in static conditions to enhance uniform gel formation. The mixture was then purified with distilled water and vacuum dried before the nano particles were calcinated. The calcination was performed at 700 °C for 2 h using a dry helium gas stream of 150 mL/min, then, at that point cooled to 150 °C under the same gas stream medium.

### 2.2. System Design

This investigation developed a mass balance model focusing on pretreatment, ethanolysis, neutralization, quenching, centrifugal phase separation, waterless purification and vacuum filtration, and vacuum distillation processes for biodiesel production (Figure 1). These mass balances were utilized afterwards to design a biodiesel plant reactor and other equipment reacquired for biodiesel production. Starting with 37 kgs of WCO, a series of pretreatments and ethanolysis

were performed. Glycerol was added to enhance the breakdown of triglycerides to diglycerides. This process lowered the acid values of the feedstock. Utilization of glycerol lowers the cost of pretreatment since the glycerol can be recovered from the biodiesel production process. The mass balances evaluate masses entering and leaving a process thereby enabling the development of correct sizing and specifications of required equipment. Detailed explanations regarding the developed processes are described in sections 2.2.1 to 2.2.7.



**Figure 1: Developed Process Flow Diagram.**

### 2.2.1. Procedure for Pretreatment

Pretreatment of FFAs was performed to lessen the contaminants and acid value of the feedstock oil and glycerolysis. Glycerolysis refers to the transesterification reaction of glycerol with triglycerides (fats/oils) to form mixtures of monoglycerides and diglycerides. This is preferred due to its lower cost. The pretreated WCO oil feed stock was premixed with CaO to enable improved ethanolysis reaction. These two compounds are immiscible hence a magnetic stirrer was utilized at 1 250 rpm for 5 min. Crude glycerol was then added during pretreatment to lower the acid value and facilitate a quicker reaction, and this reaction was performed in a batch reactor at an elevated temperature of 200 °C. Acid number was evaluated every 15 min until it dropped below 1 wt % at which point glycerol addition ceased. The process was deemed to be completed when the FFA dropped below 1 wt %.

### 2.2.2. Ethanolysis Reaction Procedure

This strategy was performed with a molar ratio of 9:1 by weight of WCO under the influence of a CaO catalyst. Ethanolysis was performed by adding ethanol to the pretreated premixture at 120 °C for 2 h. The conditions under which ethanolysis was performed are shown in Table 2. This process was stopped and mass evaluated by measuring the duration of the process (2 h).

**Table 2: Conditions for Ethanolysis**

Reaction temperature [ °C]	120
Reaction time [min]	120
EtOH-to-oil molar ratio	09:01
Mixing speed [rpm]	1250

### 2.2.4. Quenching Strategy

When ethanolysis was finished, the products were dipped in cold water of around 5 °C and this was done to suppress any possibility of reversible reaction which would subsequently lower biodiesel yields. The products of ethanolysis were

immersed in a simple reactor for 6 min and stirred at about 1 000 rpm.

### 2.2.3. Neutralization Strategy

The mixture was mixed with  $\text{H}_3\text{PO}_4$ . This process was performed at room temperatures of 20 °C and 1 atm. The duration of this process was 2 min to 5 min and was repeated until a pH of 7 was reached. The mixture was then dried at 105 °C in an oven for 8 h. This process required around 700 W. The process was deemed to be complete when pH reached 7.

### 2.2.5. Centrifugal phase Separation Process

This stage was performed to recuperate fatty acid alkyl esters (FAAE) from the products of ethanolysis. The reactants were centrifuged at 3 500 rpm for 12 min. This resulted in the formation of upper and lower layers.

### 2.2.6. Waterless and Vacuum Filtration Strategy

This process was aimed at purifying the Fatty acid methyl ester using waterless method. Waterless method is preferable to water purification due to its ability to eliminate loss of biodiesel and reduce soap formation. Ten percent Magnesium silicate ( $\text{MgSiO}_3$ ) wt % (of oil was added to the fatty acid alkyl esters (FAAE) and mixture was thereafter stirred at 1250 rpm for 6 minutes. The products of this process were then vacuum filtered.

### 2.2.7. Vacuum Distillation Process

This is the stage for recuperating a portion of the unreacted EtOH and purifying of WCOBD. The FAEE derived from the waterless purification was heated to 100 °C (slightly above boiling point of EtOH). Then, at that point a vacuum pump was utilized to extricate the EtOH fumes at 1 MPa. These conditions will eventually result in conversion of EtOH gas into liquid.

## 2.3. RESULTS

### 2.3.1. Mass balance Evaluations

CaO catalyzed ethanolysis was developed following a progression of pretreatments and ethanolysis processes until 34.04 portions of WCOBD was produced. Starting with 37 kg of crude WCO oil and following the designed processes it was found that each batch required 3.90 portions of glycerol for pretreatment and 0.74 (2 % wt of oil) portions of CaO and 3.70 (1:10) portions of EtOH for ethanolysis process. Immediately after ethanolysis the products were quenched in water baths at 5 °C for 8 min. After the quenching process, neutralization was brought about by adding 0.52 portions of  $\text{H}_3\text{PO}_4$  until pH of 7 was reached. Separation of biodiesel and glycerol is difficult because of stable emulsion formations. To improve the separation of products, an additional 9.25 (25 % wt of oil) portions of glycerol were added to the mixture. Centrifugal phase separation was performed which resulted in recovery of 41.44 portions of FAEE, 21.09 portions of glycerol and 0.51 portions of CaO catalysts. This process was followed by waterless purification and vacuum filtration in which 34.12 portions of FAEE (biodiesel) and 3.65 portions of magnesium silicate ( $\text{MgSiO}_3$ ) were blended in with the top layer (FAEE) and afterwards the combination was vacuum separated. This sub-process delivered 1.83 parts of soaps and 3.75 portions of  $\text{MgSiO}_3$  and 5.49 portions of other substances consisting of unreacted reagents represented by 'others' on the process flow diagram. The top layer comprised 34.12 parts FAEE which was further vacuum refined to deliver 34.04 parts of biodiesel (WCOBD) and 0.08 parts of EtOH.

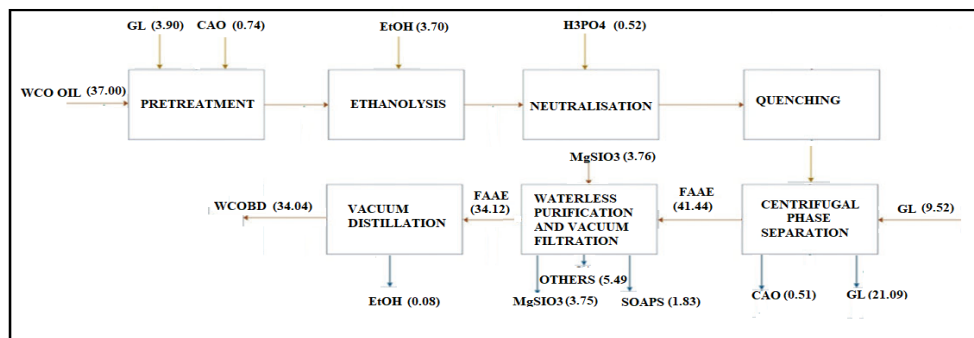


Figure 2: Material Balance Evaluated through Series of Pretreatments and Ethanolysis Processes.

Table 3: Summary of Mass Balances from Developed from Series of Experiments

Mass Balance of Wcobd Plant			
IN	WCO oil	37.00	
	Ethanol	3.70	
	CaO	0.74	
	H3PO4	0.52	
	Glycerol	13.15	
OUT	WFAAE	34.04	
	Glycerol	21.09	
	CaO	0.51	
	Soaps	1.83	
	Others	5.49	

### 2.3.2. Evaluation of Annual Production

Considering weekends and 13 holidays on the South African calendar, 248 working days are available per year. In a day two batches are possible giving a total of 68.08 kg of daily production. Multiplying the daily production and available working days gives an annual production of 16 880 kg (**16.88 tons**).

### 2.4. Economic Assessment of WCO Production

A technical evaluation of CaO catalyzed ethanolysis showed that this technique is very suitable for delivering biodiesel of high yields and quality. A high yield of 92 % (34.04/37) was unveiled by this technique. This result is similar to yields found from a comparative study (Avhad et al. 2016). Another evaluation performed by Velickovic (2016) yielded a higher volume (about 98 %) of biodiesel created under comparable reaction conditions. Nonetheless, very few studies have researched the financial assessment of delivering biodiesel from WCO oil utilizing CaO as a catalyst. In this investigation, the current best costs of materials, utilities, and costs of labor utilized was used to evaluate capital venture costs and working capital using the process model that was developed. The total equipment acquisition cost (EAC) was evaluated from the researched cost units and plant equipment shown in Table 4. Total EAC was found to be **R34,480.00** and its calculation is shown in Table 5. The value of the EAC was then utilized to evaluate most of the costs related to construction of a biodiesel plant based on the percentage allocation as per Karmee et al. (2015). Total investment capital costs for the biodiesel plant was determined by utilizing costing information acquired from suppliers in the biofuels business. These expenses were then scaled to evaluate average costs needed for standard a biofuel plant. Other local factors, for example, land value, utility expenses and work rates are aligned to the SA economy. Engineering procurement and construction (EPC) contracts from ethanol innovation suppliers were also utilized to provide a value estimation that

was utilized to consider the plausibility of the process. While this estimation can fluctuate by around 30 %, it is adequate to ascertain biofuel creation costs as they are substantially less sensitive to changes in capital expense than to changes in other factors, for example, feedstock costs. The summarized calculations are shown in Table 6. A total plant direct cost of **R83 919.27** was derived from direct costs such as equipment installation, piping, instrumentation of the plant, building and so forth. Having incorporated all the costs related to plant construction, a total capital investment of **R166 276.64** was assessed.

<b>Table 4: Researched Market Parameters Influential to Biodiesel Production Cost Evaluation</b>			
<b>Market Variables</b>	<b>Minimum Value</b>	<b>Standard Value</b>	<b>Maximum Value</b>
Biodiesel selling price (ZAR kg/1)	11.2	11.7	12.3
Feedstock oil cost (ZAR kg/1)	3.25	5.1	6.2
Cost of ethanol acquisition (ZAR kg/1)	1.6	4.8	5.3
Cost of CaO acquisition (ZAR kg/1)	1.2	1.7	3
Equipment care (% of PC)	30	75	300
Local tax (% of DFC)	230	430	650
<b>Cost of Labour (basic rate in ZAR h/1)</b>			
Operator	25	40	50
Reactor operator	35	45	65
Charge hand	40	50	80
<b>Total basic rate (ZAR h/1)</b>	<b>100</b>	<b>135</b>	<b>195</b>

**Table 5: Equipment Acquisition cost Value Evaluation**

<b>Equipment Name</b>	<b>Units</b>	<b>Unit Cost (ZAR)</b>
WCO tank (0.1 m3)	1	8,378.00
Splitter, mixer (1Kw)	1	4,698.00
Reactor (0.05 m3)	1	12,45170
Separator (decanter)	1	6,480.44
Extraction column, distillation	1	6,471.86
<b>EAC</b>		<b>38,480.00</b>

**Table 6: Capital costs for the Construction of a Biodiesel Production Plant with 16.880-ton Annual Production Capacity**

<b>Cost Categories</b>	<b>% Allocation*</b>	
Direct plant cost (DC)		
Equipment acquisition cost (EAC)		38 480,00
Equipment installation	20 × EAC	7 696,00
Refinery piping	20 × EAC	7 696,00
Instrumentation	10 × EAC	7 728,87
Insulation	3 × EAC	1 154,40
Electrical	10 × EAC	3 848,00
Utilities building	10 × EAC	3 848,00
Yard improvement	10 × EAC	3 848,00
Auxillary facilities	10 × EAC	3 848,00
Purchasing cost of unlisted equipment (UEPC)	10× EAC	3 848,00
Installation cost of unlisted equipment	50 × EAC C	1 924,00
Total plant direct cost (TDC)		83 919,27
Indirect plant cost (IC)		
Engineering	25 × TDC	20979,818
Construction	25 × TDC	20979,818
Other indirect costs		2625
Total plant indirect cost (PIC)		44584,635
Contractors' fee	5 × (TDC + PIC)	6425,1953

Contingency	$10 \times (\text{TDC} + \text{PIC})$	12850,391
Sum of contractor's fee and contingency (CFC)		3435
Total direct fixed capital cost (TDC + PIC + CFC)		131 938,91
Working capital (WC) (25 x DFC)		32 984.73
Startup and training (S &T)		1350
<b>Total capital investment cost (TDC +PIC+ CFC + WC + SVC)</b>		<b>166,273.64</b>
<b>*Percentage allocation is based on Karmee et al. (2015)</b>		

For the chosen creation process, the base selling cost of biodiesel for the business to be beneficial is ZAR10.10 /kg (673.33 US\$ /ton) at the current exchange rate of ZAR15/US\$. This cost concurs with the biodiesel selling value estimate in the Food and Agricultural Organization – Organization for Economic Co-activity and Development (FAO-OECD) Agricultural Outlook 2016–2025. Raw materials for ethanolysis constitute 56 % to the production cost which is far lower than Gebremariam and Marchetti (2019) who estimated 80 % contribution for the cost of raw materials. The calculation method followed to evaluate this cost is shown in Table 6. These results reveal that WCO is the cheapest feedstock for commercial biodiesel production since this is only 14.84 % higher than that of fossil diesel fuel.

**Table 7: Biodiesel Production Cost Based on 16 883 kg Production per Annum**

Cost Category	Calculation	Amount (ZAR) %	%
1. Raw materials	From material balance	94,780,30	55,57
2. Utilities cost	From material balance	17,140,00	10,05
Variable Costs (VC)	(1) + (2)	111,920,30	0,00
3. Maintenance	$5\% \times \text{EAC}$	1,924,00	1,13
4. Operating labor	Manning estimates	564,00	0,33
5. Laboratory cost	$30\% \times (4)$	169,20	0,10
6. Depreciation	$10\% \times \text{TDC}$	8,931.93	5,71
7. Insurance	$2\% \times \text{DFC}$	2,638,77	1,55
8. Cooperate tax	$28\% \times \text{DFC}$	36,942,90	21,66
9. Factory expense	$5\% \times \text{DFC}$	6,596,95	3,87
10. Miscellaneous	Fixed	70.00	0,04
Fixed costs (FC)	(3) + (4) + .....+ (9)	57,819.75	
Annual operating cost	(VC) + (FC)	169,740.05	
<b>Unit production cost</b>	<b>169,740.05 ZAR/16 880 Kg WFAEE</b>	<b>= 10.10 ZAR kg<sup>-1</sup></b>	
	<b>= 9.10 ZAR/ L (1liter = 0.90 kg )</b>		

## 2.5. WCOBD Plant Equipment Design and Specifications

According to Van Gerpen, Shanks et al. (2004), the easiest method of biodiesel production is to utilize a batch reactor. In this type of reactor, two charging points located at the highest point on the tank are used to introduce the reactants into the reactor tank. Nothing is retrieved or added to the reactor until the reaction is completely finished. The authors' investigation explained that this reactor tank reactor comprises a tank with an agitator and essential heating system and cooling method. The tank is a jacket vessel for minimizing heat losses. Biodiesel yield will in general increase with reaction temperature, however, excessive temperature increases beyond the alcohol boiling point (ethanol 75 °C) diminish the yield because of ethanol vaporization. Therefore, a pressurized batch reactor was selected to cater for higher temperatures.

A 50 L reactor was selected based on the sizeable feedstock supply and to lower initial construction costs and its experimental evaluation is shown in Table 7. Masses and densities of the reactants were measured, and these were utilized



to size the tank. The proportions determined from mass balances were retained and the glycerol portion was lowered to reach the 50 L capacity. Lowering glycerol from 25 % to 12.1 wt % of oil will prolong pretreatment but lower the cost of production.

**Table 8: Reactor Tank Sizing from Experimental Evaluation**

Compounds	Mass of (M)[kg]	Volume of (V)[m <sup>3</sup> ]	Density (ρ)[kg/m <sup>3</sup> ]	Experimental data [wt%]	Overall density (ρ) [kg/m <sup>3</sup> ]
WCO oil	37,00	4,16E-02	890	83.00	
EtOH	3,70	4,68E-03	790	21.12	
CaO	0,74	2,22E-04	3340	0.44	
GL	4,48	3,48E-03	1288	7.00	
<b>Total</b>	<b>45,36</b>	<b>5,00E-02</b>		<b>100,00</b>	<b>907,55</b>

The formula used to evaluate the tank masses are shown in Equations 1 to 6

- Mass of WCO oil

Mass of WCO (M<sub>wcoil</sub>) was assessed utilizing Equation 1

$$M_{wcoil} = v_{oil} \rho \quad (1)$$

- To assess the mass of ethanol (EtOH), test technique and molar proportion of 9 were used including its Table 8.

$$\text{Mass of ethanol (EtOH)} = 0.1 (M_{wcoil})(1:10) \quad (2)$$

- Mass of catalyst (CaO)

Test assessment was used to assess the measure of catalyst (calcium oxide, CaO) giving

$$\text{Mass of CaO} = 2 \text{ wt\% of } (M_{oil}) \quad (3)$$

- Evaluation of measure of added glycerol (GL)

Weight of WCO oil along with test investigation uncovered that

$$\text{Mass of added GL (MGL)} = 12 \text{ wt\% of } (M_{wcoil}) \quad (4)$$

- Total mass of reactants in the tank (MT)

$$MT = (v_{oil} \rho) + (0.1 M_{wcoil}) + (2 \text{ wt\% of } M_{wco}) + (25 \text{ wt\% of } M_{wcoil}) \quad (5)$$

- Overall density of reactants = (MT)/0.05 m<sup>3</sup> (6)

#### 2.4.1. Reactor Tank Design

According to Perry et al. (1997), the ratio of diameter to height of an unbaffled reactor should lie within the range of 0.75 and 1.5. In this design a proportion of 1: 1 is selected for use.

Reactor volume:  $V = 0.05 \text{ m}^3$  (From Table 8)

$$V_T = \frac{\pi D_T^2}{4} H_T$$

But  $H_T = D_T$

$$V_T = \frac{\pi D_T^3}{4} \quad (7)$$

Tank Diameter

$$D_T = 3 \sqrt{\frac{4 \times 0.05}{\pi}} \quad (8)$$

$$D_T = 0.4 \text{ m}$$

Considering that the  $V_T$  is 60 % of total reactor height, ( $H_R$ ) is computed as follows:

$$H_R = \frac{0.40}{0.6} = 0.663 \text{ m}$$

#### 2.4.2. Impeller Design

According to Verschuren et al. (2001), the diameter of impeller ( $D_{IM}$ ) ranges from  $(0.3 - 0.6) D_T$  and in this design a factor of 0.4 was chosen and therefore:

$$D_{IM} = 0.4 \times 0.4 = 0.16 \text{ m} \quad (9)$$

Impeller spacing equals  $\frac{D_R}{2.6918}$ , giving:

$$\text{Impeller spacing} = \frac{0.4}{2.6918} = 0.148 \text{ m} \quad (10)$$

Number of impellers equals  $\frac{D_R}{\text{Impeller Spacing}} = \frac{0.4}{0.16} = 2.5$  which translates to three impellers.

Shaft impeller was designed based on torsional stresses

$$D_{IS} = 0.208 D_{IM} = 0.033 \text{ m} \quad (11)$$

Impeller blade width and height were evaluated using the following equation

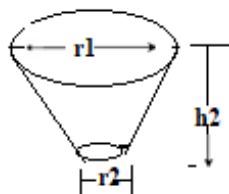
$$\text{Blade width} = \frac{D_{IM} - D_{IMS}}{4} \quad (12)$$

$$\frac{0.16 - 0.033}{4} = 0.032 \text{ m}$$

$$\text{Blade Height} = \frac{D_{IM} - D_{IMS}}{2} \quad (13)$$

$$= \frac{0.16 - 0.033}{2} = 0.064 \text{ m}$$

Clearance height from the tank bottom  $H_C$  is given by the volume of funnel shaped area ( $V_c$ ) of the round and hollow vessel (where  $r_1$  and  $r_2$  are the radii of the reactor and outlet tap, separately) was given by the Equation 9.



$$V_C = \frac{\pi h_2}{3} (r_1^2 + r_1 r_2 + r_2^2) = 0.17 D_T \pi (0.25 D_T + 0.125 D_T + 0.00625) = 0.0078 \text{ m} \quad (14)$$

Hence, clearance will be approximately 0.1 m by applying volume of conical section as the clearance.

$$H_C = 0.1 \text{ m}$$

### 2.4.3. Mixing Power Evaluation

Reynolds number describes the level of laminarity or turbulence inside the reactor (which is under 10 for laminar or more than  $10^4$  for turbulent stream systems),

$$N_{RE} = (D_{IM})^2 * \frac{N\rho}{\mu} \quad (15)$$

Whereby  $D_{IM}$  is impeller diameter 0.16m,  $N$  is impeller rotational speed 20.83 rev/s (1250 rpm) rev/s,  $\rho$  is density of reactants 907.6 kg/m<sup>3</sup> (Table 8),  $\mu$  is fluid viscosity of the oil 0.21 Pa s. Computing Reynolds number gives  $N_{RE}$  of 2304 (Liley et al. 1999).

$$P_{DELIVERED} = N_p * \rho N^3 D^5 \quad (16)$$

where  $N_p$  is the power number, which relies upon the Reynolds number calculated in Equation 15 and was acquired from nomographs as defined by Green and Southard (2019). Other variables are those defined in Equation (15). The relationship evaluated in Equation (16) is given by plots of power number versus Reynolds number for various sorts of impellers (Tilton, 1999). The power number relating to a Reynolds number of 2304 corresponding to an impeller with two blades fall within 1 and 5, giving a power range of 746 W and 3112 W. 0.746 kW was selected.

Power density of mixing =  $P_{delivered}/\text{Volume of tank } (V_T)$

$$P_{DENSITY} = \frac{0.746}{0.05} = 15 \text{ kw/m}^3 \quad (17)$$

Utilizing the discoveries by McCabe, Smith et al. (1993) for power density must be above of 2.0 kW/m<sup>3</sup>, the computation for Equation (17) showed that effective blending is possible under these conditions.

**Table 9: Summary of Designed Equipment Specifications**

Parameter	Designed Value	
Tank capacity (L) or [m <sup>3</sup> ]	50	
Total tank height [m]	0.663	
Tank diameter [m] for reaction	0.4	
Tank height [mm] liquid/ solid fill	388	
Liquid/ solid fill for reaction [% ]	60	
Impeller diameter[mm]	160	
Impeller blade width [mm]	32	

Impeller blade length [mm]	50	
No. of impellers	3	
Impeller location height [mm]	100	
Diameter of impeller shaft [mm]	33	

### 5.3 CONCLUSIONS

This research work was aimed at designing a novel process flow based on CaO catalyzed ethanolysis process. From process design assessment, the evaluation affirmed that the establishment of a mini pilot plant by small to medium fuel producers in South Africa is achievable. However, the reaction process displayed high energy consumption owing to CaO thermal decomposition and pretreatment at raised temperatures to enhance the ethanolysis process. Nevertheless, it has been affirmed that biodiesel creation by means of CaO catalyzed ethanolysis is a novel way to enhance biodiesel from WCO. The developed technique, though challenging in terms of the many sub-processes, produced biodiesel of high quality, with a lower environmental impact as waterless purification was utilised. In terms of economic assessment this production process seems viable as the production cost per ton lies within the FAO anticipated production cost per ton. From mass balances a 50 L reactor was successfully designed together with some of its key auxiliaries.

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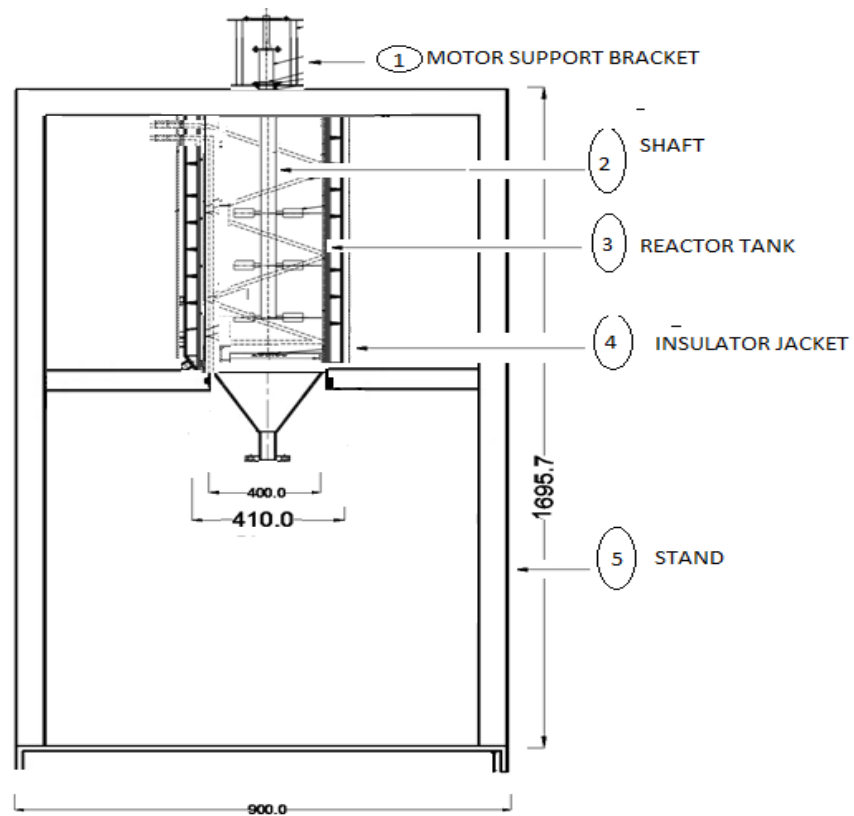
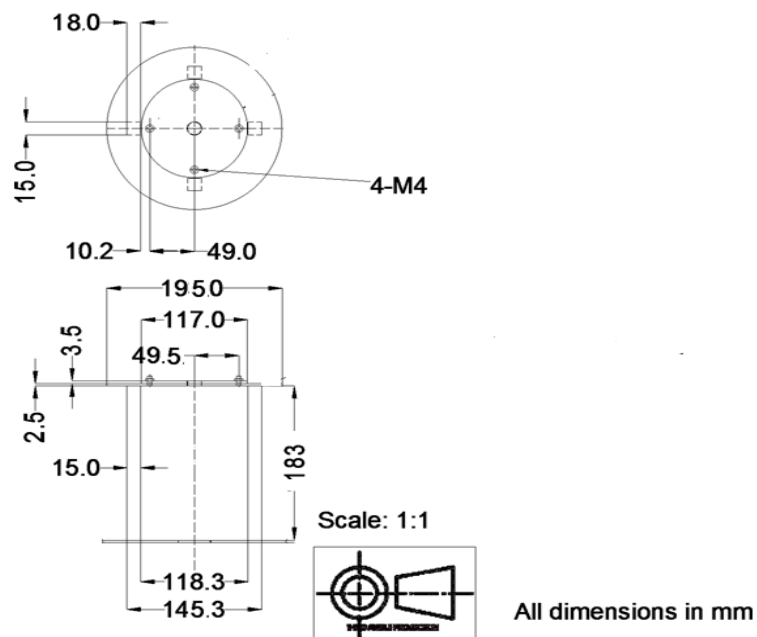
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## APPENDICES

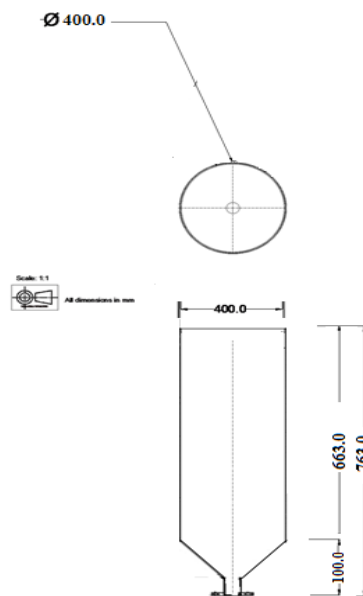
### APPENDIX 1: DETAILED MASS BALANCE

Materials and Utilities		Batch Process
		kg/batch
Final Product	WCOBD	34.04
Raw material	WCO oil	37.00
	Feed of EtOH (Ethanolysis)	3.70
	CaO (Ethanolysis)	0.72
	Feed of GL (separation)	9.25
	H <sub>3</sub> PO <sub>4</sub>	0.52
	Feed of MgSiO <sub>3</sub>	3.76
Byproduct	Reusable GL	21.09
	Recyclable EtOH	0.08
	Reusable CaO	0.51
	Waste soaps	1.83
	MgSiO <sub>3</sub>	3.75
Utilities	Pre-treatment	
	Heating [MJ/batch]	8,480.00
	Stirring (mechanical)[MJ/batch]	132.00
	<b>Reaction</b>	
	Heating [GJ/batch]	9,270.00
	Stirring (mechanical) [MJ/batch]	154.00
	Centrifugal separation [MJ/batch]	68.00
	<b>WCOBD purification</b>	
	Mechanical stirring [MJ/batch]	26.32
	Vacuum filtration [MJ/batch]	19.10
	Vacuum distillation (heating) [MJ/batch]	25.10

**APPENDIX 2: ASSEMBLED PRESURED BATCH REACTOR****APPENDIX 3: DETAILS OF MOTOR SUPPORT BRACKET**

## APPENDIX 5: SHAFT DESIGN



**APPENDIX 6: REACTOR TANK****APPENDIX 7: INSULATOR JACKET**